Technical and Economic Assessment for Biomass-to-Ethanol

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Abstract

NREL has studied biomass-to-ethanol process designs and their economics and the results are currently used to help direct research and understand potential market penetration issues. NREL's work has also been used as a starting point for corporate partners in their technology development and process design work. Using a process and parameters that are believed possible in a plant that can be built in 2010 and n^{th} plant modeling techniques, the resulting minimum ethanol selling price (MESP) is \$1.09/gal ethanol. Our investigations show that a pioneer, or first of a kind, plant using the same parameter set would have an MESP between \$1.53/gal and \$1.75/gal (range includes one standard deviation around mean). Feedstock variability and Monte Carlo analyses are also reported and show ranges around the \$1.09/gal result. We propose the following near-term work items: understanding process emissions requirements and process designs to achieve those requirements, improved tracking of feedstock components, modeled solid/liquid separation parameters that match experimental data, and better modeling of overliming. Long-term work is also proposed.

Introduction

During the last 25 years, NREL has been studying process designs and options and estimating costs of fuels developed from biomass (for examples see Wright 1988, Hinman et al. 1992, and Wooley et al. 1999a and 1999b). The economic estimates serve three primary purposes. The first is to estimate an absolute cost of bioethanol. That information is then used to judge bioethanol's potential for market penetration. The second purpose is to understand the process and economic impacts of proposed research strategies and consequently to guide process development. By translating research goals into process performance parameters that can be reflected in our process models, we can predict the impacts of these goals on the bottom line – cost savings per unit of product. Finally, current and projected technology costs are used as inputs to policy discussions.

While studying process options and economics, NREL has developed tools that are available for commercial and academic research and development. These tools include techniques for modeling within ASPEN+ and physical property model parameters that are being used by corporate and academic partners in their own studies. Corporate partners have used the tools and techniques as starting points to determine the economic viability of technology they are interested in developing. Research partners use the tools to help select areas where research can provide a significant impact on process economics.

Approach to Process Design and Economic Analysis

Technical and economic assessment demands a continuously improving understanding of the process. The process understanding begins with development of process flow diagrams, which were initially developed by NREL in conjunction with outside engineering companies. Once process flow diagrams are available, material and energy balances are calculated. Initially, those calculations were done by hand or using spreadsheet software. We now use the ASPEN+$^{®}$ chemical process modeling software developed by Aspen Tech. (Cambridge, MA) because of the increased rigor and reduced potential for errors. Experimental material balance tools have been developed for research-relevant areas so that the most complete empirical results can be easily incorporated. Information from ASPEN+ material and energy balances is used to estimate capital and operating costs. Those costs are estimated from various sources including ICARUS cost estimation software (also sold by Aspen Tech.) for standard equipment items, vendor cost quotes for unusual items, and engineering company estimates for package areas like wastewater treatment.

Finally, a discounted cash flow rate of return (DCFROR) calculation is used to convert capital and operating costs to a minimum ethanol selling price (MESP). The MESP is the minimum price the market can bear while the ethanol production facility makes its required return on investment – 10% after taxes for our analysis. If the market's price is higher than the MESP, the facility makes a larger profit. If it is lower, the facility makes a smaller profit (making it less attractive to investors) or a loss.

Previous NREL Work

In 1991, NREL developed a complete process design, estimated design parameters that could potentially be achieved, and estimated economics of a biomass to ethanol process (Hinman et al. 1992).

That work was reviewed and updated over the past decade. In 1999 an update was reported in an NREL publication entitled "Lignocellulosic Biomass to Ethanol Process Design and Economics Utilizing Co-current Dilute Acid Prehydrolysis and Enzymatic Hydrolysis Current and Futuristic Scenarios" (NREL/TP-580-26157 (http://www.afdc.doe.gov/pdfs.3957.pdf). The parameters used for this study were believed achievable in the middle of the first decade of the 21st century and resulted in a MESP of $1.44/gal.

Last year's ESP gate 3 review further updated that work. For that review, a process model was developed with corn stover as the feedstock and parameters that were believed achievable in the 2005-2008 timeframe. The model had a MESP of $1.30/gal and many sensitivities were run on that case to direct research by the project and help identify situations where production of bioethanol could be economically viable in the near term.
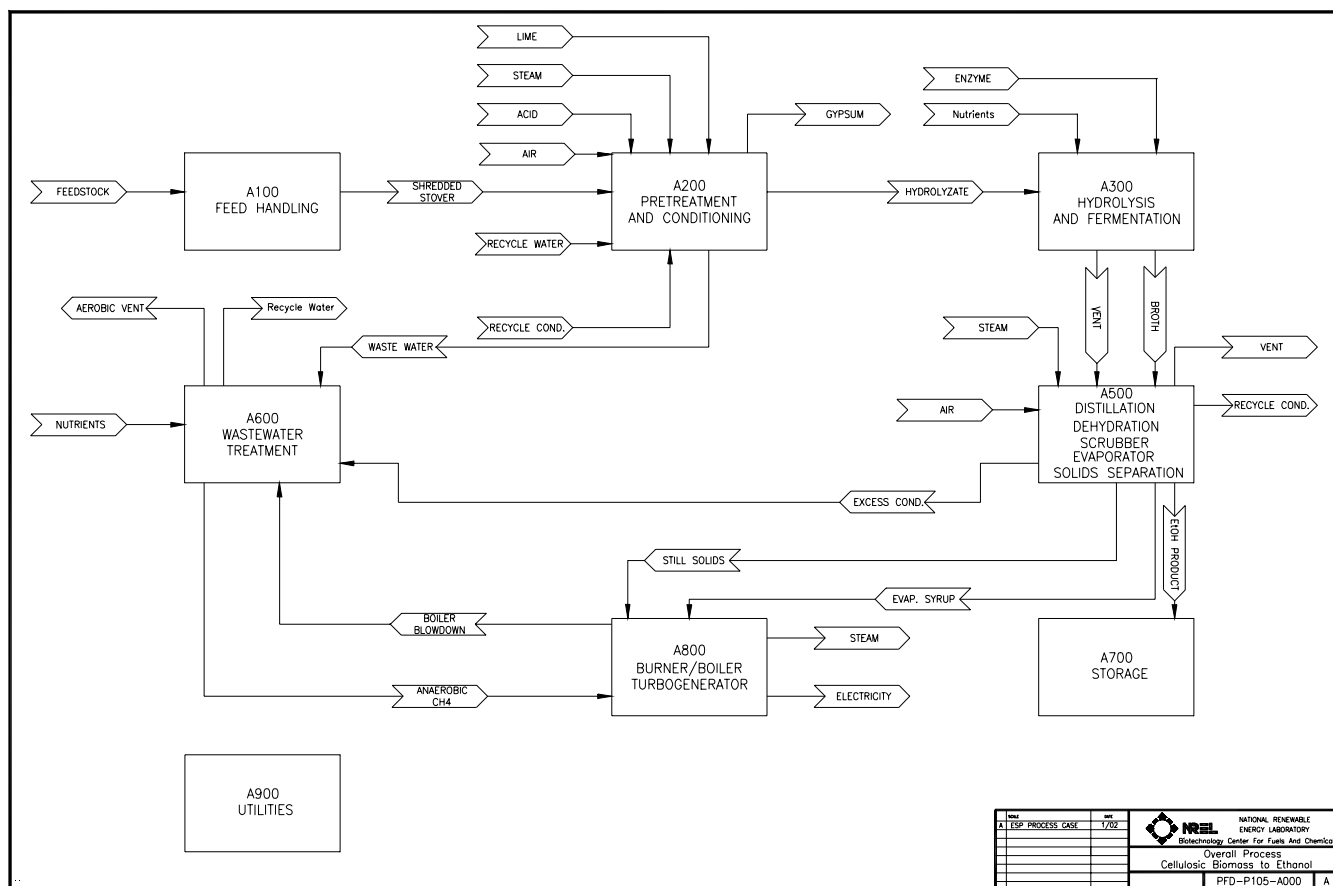http://www.ott.doe.gov/biofuels/pdfs/stage2_overview.pdf

Figure 1. Overall Process, PFD-P105-A000

In 2002, the process design was further updated to the one shown in Figure 1. The parameters chosen for that analysis are believed to be achievable by 2010. The new timeframe was chosen because ESP is no longer developing technology that can be commercialized in the near future but instead is investigating the long term potential of the process. The results were published in another NREL report http://www.ott.doe.gov/biofuels/pedownload.html#6483 The MESP resulting from that work is $1.07/gal ethanol.

This year, further model corrections, physical property improvements, and switching the cost year from 2000 to 2001 to match the current annual energy outlook (http://www.eia.doe.gov/oiaf/aeo/) increased the MESP to $1.09 for the current process model.

Process Design

For analyses since 1999, the design is divided into nine sections (see Figure 1): Feedstock Handling, Pretreatment and conditioning, Enzymatic Hydrolysis, Fermentation, Ethanol Purification and Recovery, WWT, Storage, Lignin Utilization, and Utilities.

The design is based on a 2000 dry metric tonne/day (biomass) facility. The estimated cost of corn stover is $30/dry ton at the plant gate. Although corn stover is a variable feedstock, one composition had to be assumed for the base case. That composition is shown in Table 1. Compositional variability and its effect on process economics is discussed below.

Table 1. Feedstock Composition

Component	% Dry Basis
Glucan	37.4
Xylan	21.1
Lignin	17.5
Ash	6.1
Unknown Soluble Solids	5.6
Acetate*	2.0
Protein	4.0
Extractives	3.3
Arabinan	2.2
Galactan	1.2
Mannan	0.2
*Acetate is the acetate groups present in the hemicellulose polymer. They are generally converted to acetic acid in the prehydrolysis reactor.	
Moisture	15.0%

The plant receives round corn stover bales on truck trailers. As the trucks are received, they are weighed and unloaded by forklifts. From there, the bales are conveyed to an automatic

unwrapping system that cuts away the plastic wrapping and/or net surrounding the bales. The unwrapped bales are conveyed to a wash table, which both breaks up bales and washes dirt and grit from the corn stover. The washed stover is then conveyed past a metal detector to remove metallic impurities, after which it is introduced to primary and secondary shredders. Finally, the washed and milled stover is conveyed to prehydrolysis. Including the capital and operating costs of preparing stover for prehydrolysis, the cost of corn stover at the prehydrolyzer is $35.60/dry ton as compared to $30.00/dry ton when it is received.

Pretreatment is necessary to break up the integrated fibrous structure of the lignocellulosic biomass, making the carbohydrates more accessible to enzymatic conversion. Most of the hemicellulosic portion of the feedstock is converted to soluble sugars, primarily xylose, mannose, arabinose, and galactose. This conversion is accomplished using dilute sulfuric acid (1.1% w/w) and high temperature (190°C) in a brief time (2 minutes residence time). The xylan conversion yield (85%) has been achieved at NREL but is at the upper edge of the data presented in the pretreatment section of this document. These conditions also solubilize some of the lignin in the feedstock and liberate acetic acid from the hemicellulose. Degradation products of pentose sugars (primarily furfural) and hexose sugars (primarily hydroxymethyl furfural (HMF)) are also formed. The reactor is designed to operate at high solids concentrations and is currently modeled at 30% total solids. Based on metallurgy testing done through Harris, the reactor materials of construction (MOC) are specified to be Incoloy 825-clad steel for all wetted parts.

Following the pretreatment reactor, the hydrolysate (liquid and solids) is flash cooled, vaporizing a large amount of water, much of the furfural, and a portion of the acetic acid. Removal of these is beneficial as they can be detrimental to downstream fermentation. The hydrolysate undergoes solid/liquid separation using a pneumatic pressure filter (Pneumapress). Vendor testing has shown that dry cakes, near 55% insoluble solids (IS) with 99.5% IS recovery, are possible with this unit; however pilot scale information is not yet available.

We assumed that the filtrate requires conditioning by "overliming." In this step, lime is added, raising the pH of the stream to 10. Steam is also directly injected to heat it to 50°C. Neutralization and precipitation of gypsum follow the overliming step. The gypsum is filtered out and the hydrolysate is mixed with the solids and dilution water before being sent to enzymatic hydrolysis and fermentation.

The conditioned filtrate and hydrolysate solids are remixed for hybrid hydrolysis and fermentation (HHF) to convert cellulose to glucose and ferment the sugars that are present. Both Genencor Intl. and Novozymes Biotech are developing improved enzymes for biomass conversion. HHF was chosen because those improved enzymes are expected to have faster kinetics at high temperatures; however, those enzymes are also expected to be slowed down when sugars build up similar to current enzymes. In the process model, purchased cellulase enzyme (estimated to cost $0.10/gal EtOH) is added to the hydrolysate and held at 65°C for approximately 1.5 days. During this time, most of the enzymatic hydrolysis will take place. The temperature is then cooled to 37°C (assuming that a thermotolerant fermenting organism is not used) and a "yet-to-be-determined" recombinant fermenting organism is added along with the required nutrients. The fermentation broth is then held for approximately 2 days during which the enzymatic hydrolysis concludes and the co-fermentation of glucose, xylose, and arabinose

takes place. The designed case results in an ethanol concentration in the resulting "beer" to be just over 5%.

Product recovery involves distilling the beer to separate the ethanol from the water and residual solids. A mixture of nearly azeotropic water and ethanol is purified to pure ethanol using vapor-phase molecular sieves. Solids from the distillation bottoms are separated (via Pneumapress) and sent to the biomass burner/boiler/turbogenerator. The distillation bottoms liquid is concentrated by evaporation, using waste heat. The evaporated condensate is returned to the process and the concentrated syrup is sent to the burner to minimize the load to wastewater treatment.

Part of the evaporator condensate, along with other wastewater, is treated by anaerobic and aerobic digestion. During anaerobic digestion, 90% of the organic material present is converted to methane and carbon dioxide (biogas). The biogas from the anaerobic digestion is sent to the burner for energy recovery. In the aerobic digestion lagoon, another 90% of the remaining organics are removed. The treated water is considered suitable to recycle and is returned to the process. Aerobic sludge is clarified, filtered, and sent to the burner/boiler for disposal.

The solids from distillation, the concentrated evaporator syrup, and biogas from anaerobic digestion are combusted in a fluidized bed combustor to produce steam for process heat. The majority of the steam demand is in the pretreatment reactor and distillation areas. A multistage turbine and generator are used to generate electricity from the excess steam for use in the plant and for sale to the grid. Other utilities provided include cooling water, chilled water, plant and instrument air, process water, and clean-in-place (CIP) system.

Results

In the 2002 design case (Aden et al. 2002), the parameters were estimated as what may be possible in a plant that could startup in 2010. That parameter set resulted in an MESP of $1.07/gal (2000$). With the exception of the costing year, those parameters were not changed for this analysis and the resulting MESP is $1.09/gal (2001$). A 2000 dry metric tonne/day facility with the modeled parameters will produce over 69,000,000 gal ethanol annually at a yield of 90 gal/dry ton corn stover. The estimated total project investment (TPI) is $201,000,000 (2001$) so the capital cost per annual gallon is $2.90. This estimate is based on an nth plant analysis and is accurate to +30%/-15% of capital.

Figure 2 shows the cost breakdown for each section of the facility. Raw materials are the most important cost element with biomass feedstock cost at 31% and cellulase enzyme at 9% of the $1.09/gal MESP. The primary plant cost areas are pretreatment/conditioning (19%), distillation and solids recovery (12%), and saccharification and fermentation (8%).
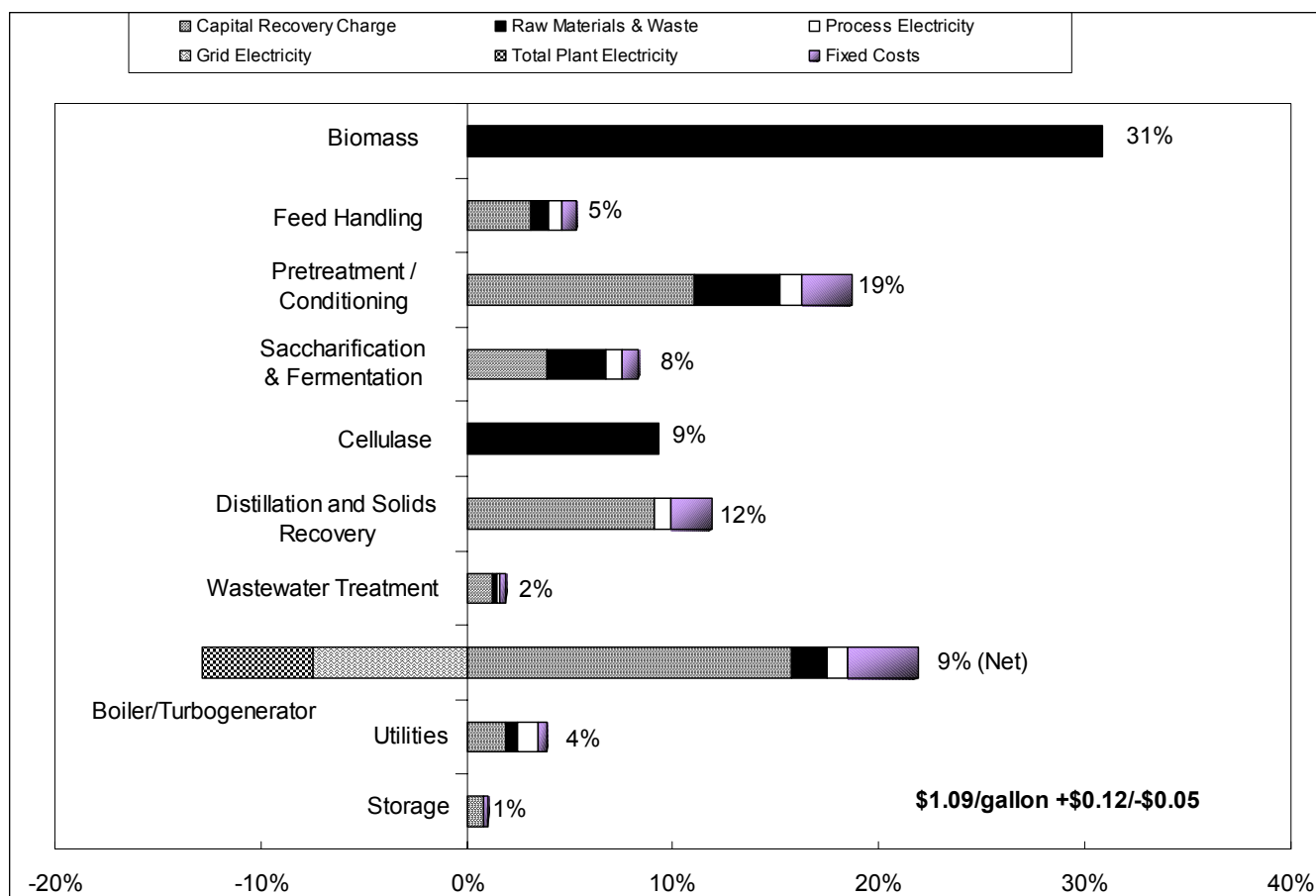
Figure 2. Process cost breakout by process areas

Cost of Sugar

In a biorefinery, slipstreams may be removed from the ethanol process to produce other products. Two of the potential sugar slipstreams are hydrolysate liquor immediately after pretreatment and conditioning and saccharified slurry before fermentation. Minimum selling prices for the sugars in each stream were calculated to keep the MESP constant with reduced ethanol production. The hydrolysate contains 9% (wt%) soluble sugars (monomeric and oligomeric) and that sugar stream requires a minimum selling price of $0.055-$0.063/lb sugar depending upon the size of the slipstream. The saccharified slurry contains 14% (wt%) soluble sugars (monomeric and oligomeric) and that stream requires a minimum selling price of $0.057-$0.064/lb sugar. The enzyme cost contribution of that minimum sugar selling price is $0.007/lb sugar.

Pioneer Plant Costs

The design reported here uses an nth plant analysis to determine all of the economics. In 1981, The Rand Corporation attempted to link the results from this type of analysis to the actual cost of building and starting-up a first plant (Merrow et al. 1981). They broke the first plant effects into cost growth (increase in TPI) and plant performance during months 7-12 after start-up. We used the techniques proposed by Rand and Monte Carlo analysis to calculate the potential TPI and MESP for a first plant. An average TPI of $378,000,000 was calculated with a standard

deviation of $38,000,000. An average MESP of $1.64/gal was calculated with a standard deviation of $0.11/gal. These costs are much higher than the nth plant costs reported above and emphasize the need for more experimental mass balance closure, process integration, pilot and demonstration scale testing, and improvement of thermodynamic models.

Analysis of Multiple Stover Compositions

Each of the 738 different stover compositions collected and measured by NREL (Thomas et al. 2002) were normalized to 3.7% non-structural sugars and simulated. Normalization was necessary to account for possible removal of non-structural sugars in feedstock processing (i.e., milling, washing, drying). Only 3 of the 738 different compositions caused simulation errors that prevented calculation of the MESP. Also, 151 of the 735 cases did not have sufficient steam; however, their economic results are reported because modification of the steam system or additional combustion feeds will only affect the economics in a minor way. That minor effect was considered unnecessary for modifications in sensitivities but the changes would have been made if any of the cases were being used for a new "base case."

Figures 3 and 4 show the variation of MESP over the population. The mean is $1.14/gal with a standard deviation of $0.05/gal. One can easily see that the middle 90% of the samples lie between $1.07/gal and $1.24/gal.
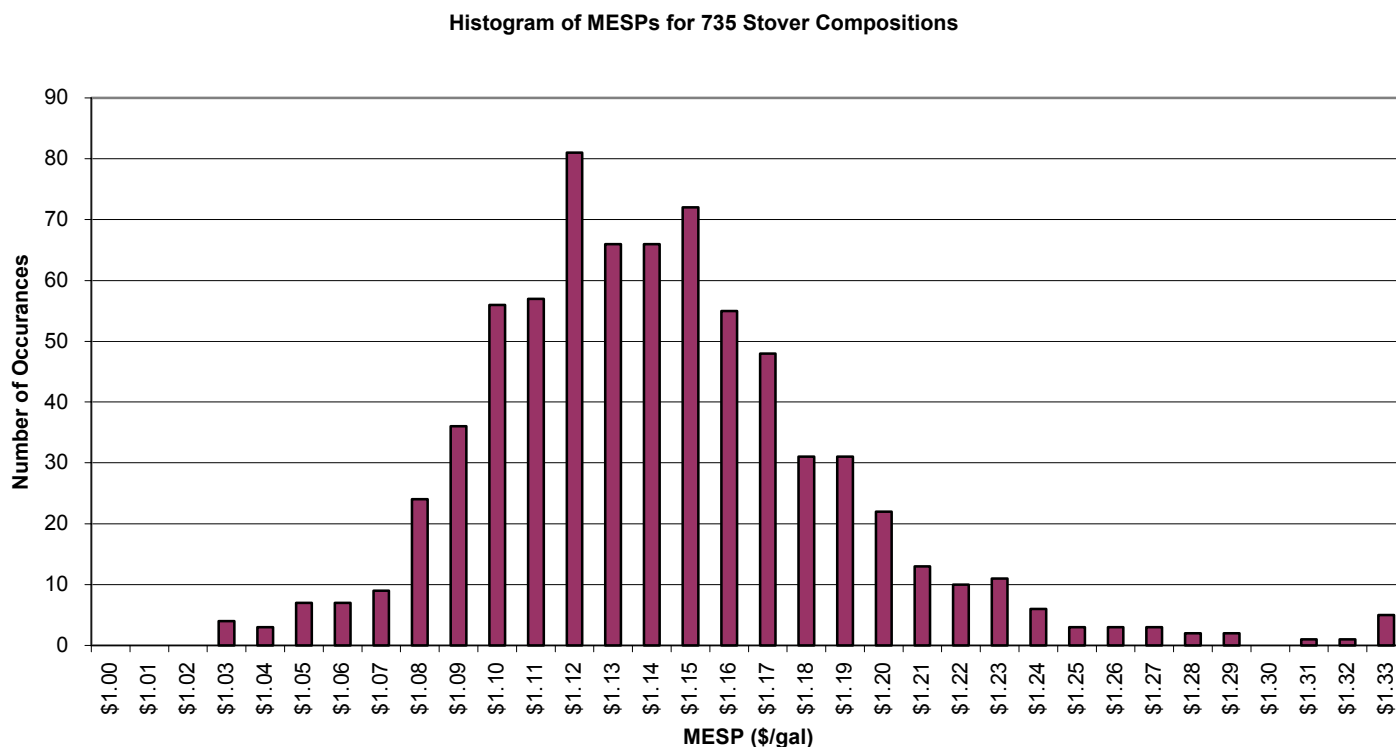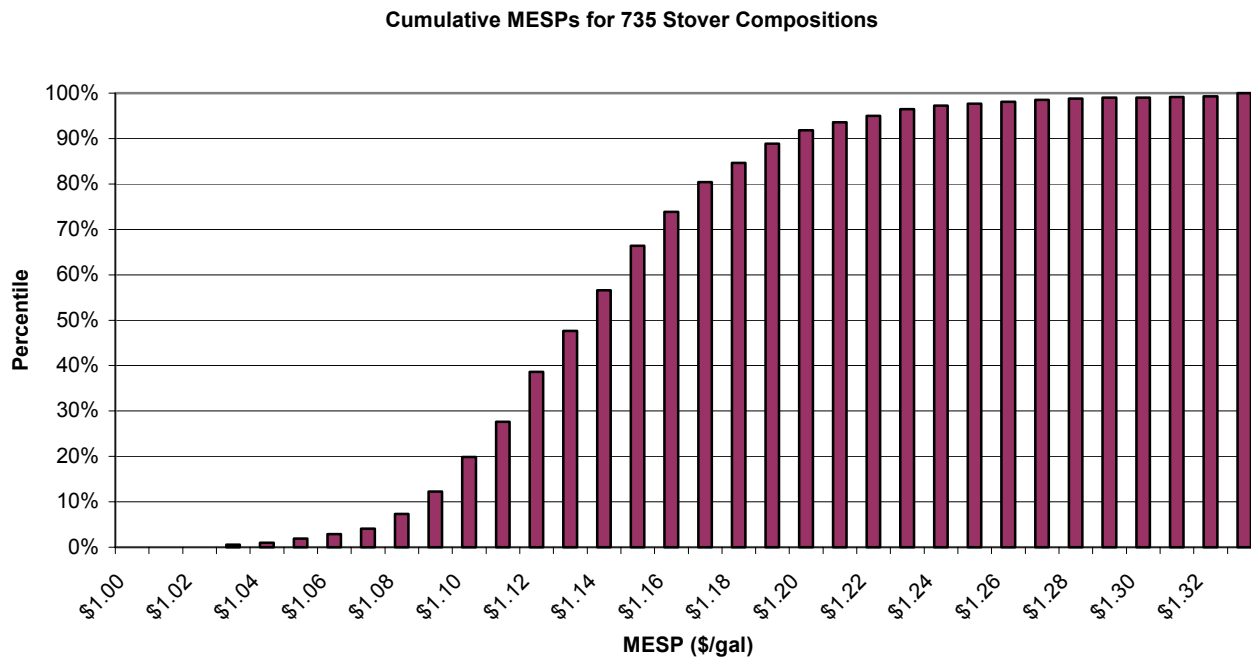
Histogram of MESPs for 735 Stover Compositions

Figure 3. Histogram for 735 stover compositions (R0302C)

Cumulative MESPs for 735 Stover Compositions

Figure 4. Cumulative MESPs for 735 stover compositions (R0302C)
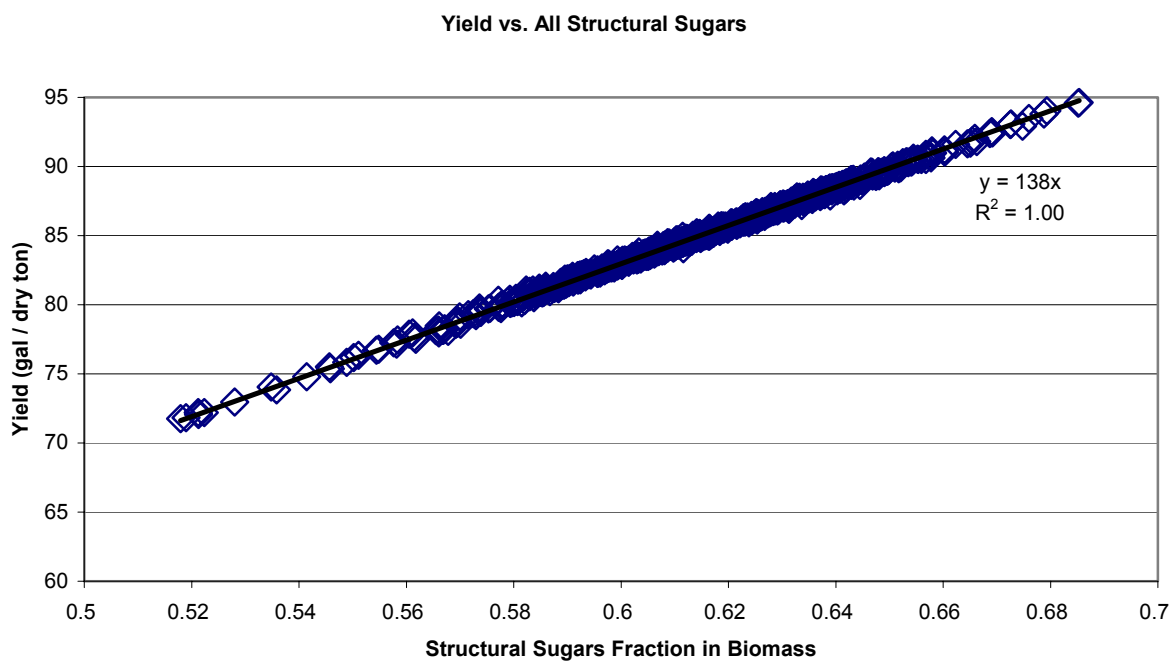
Yield vs. All Structural Sugars

$y = 138x$
$R^2 = 1.00$

Figure 5. Yield as a function of structural sugar fraction for 735 stover compositions (R0302C)

MESP vs. All Structural Sugars

The chart plots MESP ($/gal ethanol) on the y-axis (from $1.00 to $1.40) against Structural Sugars Fraction in Biomass on the x-axis (from 0.5 to 0.7). A power fit curve is labeled:

$$y = 0.7155x^{-0.9592}$$
$$R^2 = 0.9854$$

and a linear fit is labeled:
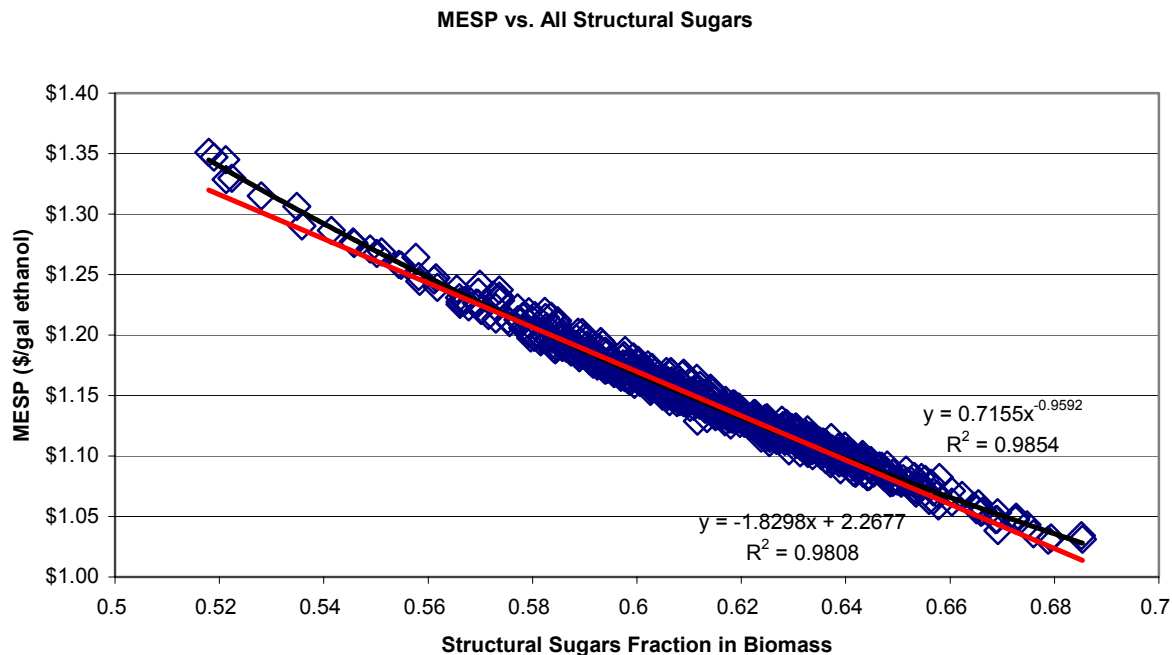
$$y = -1.8298x + 2.2677$$
$$R^2 = 0.9808$$

Figure 6. MESP as a function of structural sugar fraction for 735 compositions (R0302C)

Figures 5 and 6 show the yield and MESP as a function of the fraction of the biomass that is structural sugars. Figure 5 indicates that each percent change in the feedstock's structural sugars will modify the yield by 1.38 gal/dry ton. Figure 4 shows the effect of structural sugars on MESP. Scatter within the figure is due to non-carbohydrates within the feedstock (i.e., if two feedstocks are identical in carbohydrate composition but one has a higher percentage of lignin it will have a lower MESP because of the energy available in the lignin that would not be in other fractions). If a linear curve fit were used in figure 4, it would indicate that a 1% change in structural sugars would modify the MESP by $0.018/gal.

Confidence Analysis

Like experimental results, a single economic result is not as useful as a range showing the confidence interval around that result. In modeling work, from environmental studies to the business predictions, Monte Carlo analysis is often used to calculate that range. Monte Carlo analysis selects random numbers within defined functions for parameters, solves the model, and repeats the process numerous times to predict the modeled system's uncertainty.

Several software programs are available to simplify this work; we used Crystal Ball, which is an add-in for Microsoft Excel. Crystal Ball allows a user to define functions for parameters and then solves the Excel workbook with those parameters and allows the user to define Excel VBA code that will run during the Monte Carlo study. We developed a technique and code for that technique that transfers parameters from Excel to a text file, starts the ASPEN+ batch simulation from Excel (using DOS batch files), reads parameters from the text file into ASPEN+, solves the

ASPEN+ model, and reads results into Excel. That process coupled with use of Crystal Ball allows for Monte Carlo analysis with our techno-economic models.

For the initial analysis, feedstock parameters and several high-impact process yields were varied. The feedstock composition functions were defined to match the 738 compositions described above. The feedstock parameters varied follow: cellulose fraction, hemicellulose fraction, lignin fraction, and fraction of hemicellulose that is xylan. The remaining hemicellulose is proportionately split into galactan, mannan, arabinan. Also, the remaining biomass is proportionately split into ash, acetate, protein, and soluble solids. Cellulose and lignin were the only two major components that correlated so they were correlated for the Monte Carlo analysis.

The high-impact process yields that were varied follow: xylan to xylose yield in prehydrolysis, cellulose hydrolysis yield in enzymatic saccharification, and ethanol yields from all 5 fermentable monomeric sugars.

Over 1000 simulations were run with random parameters as defined by their functions. The resulting MESPs for those simulations are shown in figures 7 and 8 with an average value of $1.21/gal and a standard deviation of $0.06/gal. Most of the results are above the base-case's $1.09/gal MESP because most of the feedstock compositions are lower in total structural sugars than the base case values and some of the yield functions used to model the high-impact parameters result in over 50% of them being less than the base case values.
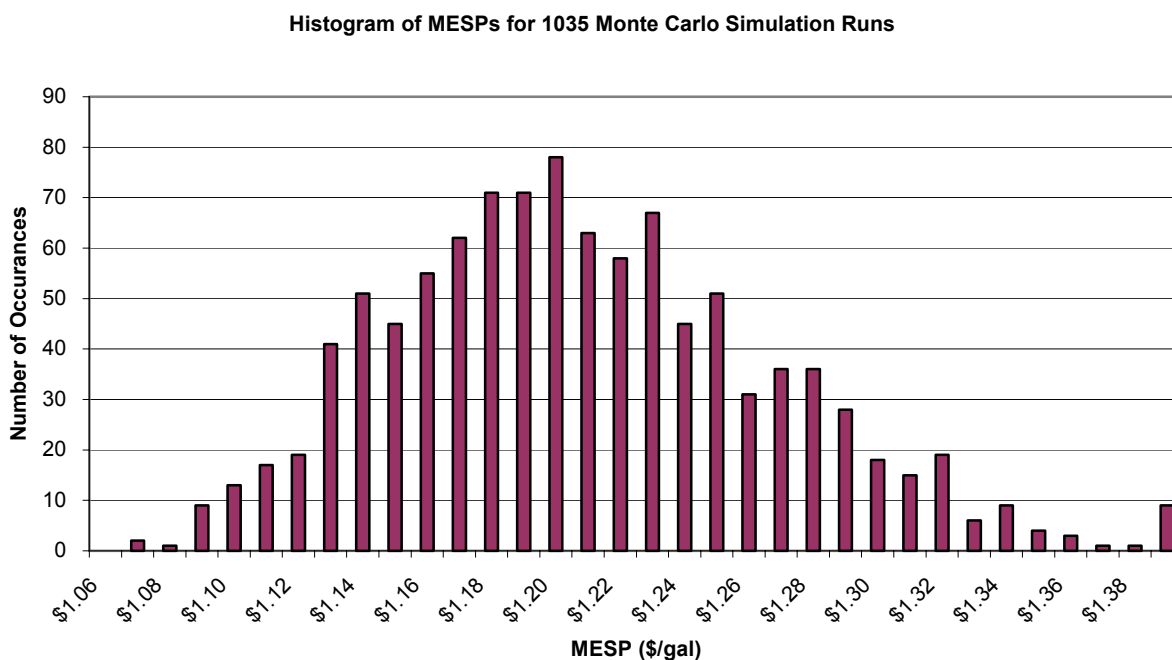
Histogram of MESPs for 1035 Monte Carlo Simulation Runs

Figure 7. Monte Carlo MESP Results Histogram (R0302D)

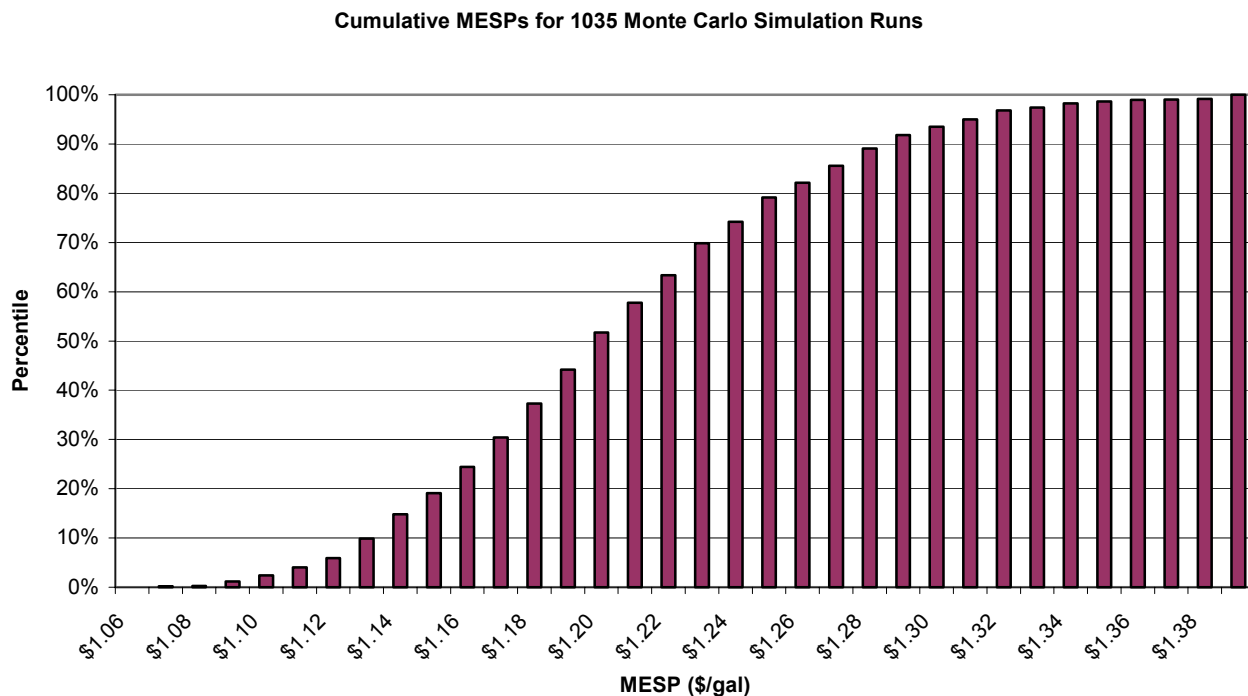**Cumulative MESPs for 1035 Monte Carlo Simulation Runs**

Figure 8. Monte Carlo Cumulative MESP Results (R0302D)

Proposed Work

We propose near-term work to further improve understanding of this biomass-to-ethanol process and to improve the overall model. The proposed work areas include facility emissions and feedstock handling, feedstock composition, solid/liquid separation, and overliming.

The Harris Group Inc. has been contracted by NREL to review all emissions and biomass handling in the biomass-to-ethanol process design. The emissions include gaseous emissions to the atmosphere from the scrubber and the boiler, wastewater that needs further processing, and solid waste emissions from both overliming and the boiler. Harris is investigating permitting requirements for all emissions and potential process changes (e.g., better water management) and treatment requirements. Harris is also investigating biomass harvest and storage issues to improve the biomass handling area design to better match the most likely harvest and handling scenarios. The results from this contract will be included in the biomass-to-ethanol process design.

As shown above, corn stover's composition varies. The carbohydrate composition used for standard designs has traditionally been an average of several samples but now that many samples are available a standard composition will be chosen. That standard carbohydrate composition will be paired with an elemental composition of the same material and the material's energy content so that the two will match within the model. We will also attempt to improve tracking of components through the process. For example, all of the protein in the feedstock is assumed to

be converted to a water-soluble form in pretreatment but that is probably not the case and an improved tracking will improve decisions made with model input.

The current parameters for solid/liquid separation were calculated from bench-scale tests for the Peumapress. Now that a pilot-scale unit is at NREL, improved parameters can be measured to verify the applicability of the design and separation efficiency.

Overliming is necessary to condition the hydrolysate and reduce its toxicity for many fermentation organisms; however, the actual reactions and products' solubilities are poorly understood. We hope to take information from overliming experiments and their material balances to improve the ASPEN+ model. We expect to add new components to the model, improve the separation parameters, and compare the solubility measurements with those predicted using electrolyte property models.

On a longer term basis, we propose work in developing LP models for biorefinery optimization, addition of kinetic models to the process models to improve the sensitivity analyses by linking yields to conditions, further risk analyses, and keeping the model updated with additional understanding gained by scale-up experiments.

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